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Deliverable report

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Summary

Task 12.3 aims to investigate a CO_2-capture process integrated with a wood-fired power plant (CHP) using an absorption system. The first step involved understanding the basics of the absorption process to be used. An important element in this work is to evaluate the energy requirements of the process and district heating network of the wood-fired power plant under consideration. The evaluation showed that a large amount of thermal energy remains unused in summer and must be re-cooled. This unused thermal energy can be used to operate the CO_2 absorption plant.

Three different absorption plant sizes were investigated. The validated simulation of the CO_2 absorption plant can be found in this report. Using this simulation model, the operating parameters were determined and different process optimization variants were tested. The next step was to use the mass transfer model in order to determine the operating ranges of the plants as well as for cost calculations and for calculating the annual operating data. A technical-economic comparison with various development scenarios and a sensitivity analysis round off this work. In the final recommendation for action, it is clear that the operation of the plant is economically dependent on many boundary conditions. In particular, the CO2 sales price as well as the transport and final storage cost have a significant influence. Also, changes in governmental policy conditions must be taken into account.

1 Introduction

In view of the policy objective of net zero carbon emissions, the deepest possible level of decarbonization must be achieved in all sectors. However, the decarbonization of the industry sector is challenging due to the high complexity of the systems, heterogeneity, and high-temperature levels. In addition, the emissions due to process emissions resulting from chemical or physical reactions are hard to abate. Carbon capture, utilization, and storage (CCUS) technologies can play a crucial role to meet the Swiss goal of net-zero greenhouse-gas emissions by 2050. The net-zero target can only be achieved if CO_2 is captured from either point sources or from the atmosphere and stored (CCS) permanently. In some industrial companies in Switzerland, the required process heat is at temperature levels unreachable by solar, geothermal sources, or heat pumps. Such processes require fuel combustion. However, fossil fuels will likely be untenable in a net-zero scenario unless the resulting CO_2 is captured and stored. If fuels are biogenic in origin, capturing and storing CO_2 can generate negative emissions, i.e. net removal of CO_2 from the atmosphere. The bulk of negative emissions can be created with thermal plants that burn biogenic carbon-containing fuel such as wood or waste. In Switzerland, these are mainly waste-to-energy plants, wood combined heat and power (CHP) plants, and cement plants. Nevertheless, not all CO_2 point sources contain biogenic fuel such as the production of cement – where could be used as an industrial point source for the collection of CO_2. In the long term, it can be expected industrial processes to be powered by renewable energy, whether through the electrification of processes or fuel switch to renewable fuels (see Energy Perspectives (Bundesamt für Energie, 2020)).

1.1 Objective

T12.3 aims to understand how to best integrate CO_2 capture to an existing wood-fired power plant. The analysis will examine, techno-economic of CO_2 capture from the flue gas and how best to use the excess thermal energy from the CHP. The task is to identify the optimal design of the plant in terms of its operating range based on the annual operation of the wood-fired plant. Therefore, process integration and optimization are key to this work. In addition, different plant variants are simulated and evaluated various criteria. Integration of the CO_2 is conducted using Pinch Analysis (PA), where heat recovery measures are also recommended. Simplified assumptions are made in the economic analysis, the transport, final analysis to take into account transport, final storage, and the changing boundary conditions. Through this, a recommendation for action can be for the next steps can be given.

1.2 Procedure

This deliverable first introduces the basic technological process of CC, where the scope focuses more on the absorption process. In addition to that, the operating parameters of the absorption process are identified. Section 3 introduces the CHP case study that is integrated into the CC process. For the case study, a suitable scrubbing agent is identified and the integration to CC to the CHP was investigated. The characterization of the CHP is followed by an outlook on its operating modes and the feed-in tariff. With the software CHEMCAD, different simulation models are created and reported in Section 4, where three different configurations are evaluated. With the help of these data of the absorption plants, the annual operating data of the data of the CHP, the process data and the district heating data, annual calculations could be made. In addition, evaluations in CHEMCAD and estimation formulas from the literature the annualized investment and operating costs are calculated. For the development of transport and final storage costs, two scenarios are considered. Possible changes in CO_2 and electricity prices and their influence on the economic viability of the plant are examined with the aid of a sensitivity analysis. The work is completed by showing possible CO_2 negative emissions, investment, and operating costs as well as recommendations for further action.

2 Basics in carbon capture

In the technological solutions, CO_2 is either captured directly from the atmosphere, which is known as direct air capturing (DAC), or from a CO_2-enriched flue gas (point source), which is e.g. extracted from a waste incineration plant (MSWI) or a combined heat and power (CHP) plant. Separation from an enriched gas is technologically simpler and more efficient than in the case of capture from the atmosphere (DAC). CO_2 capture from enriched streams traditionally were used mainly for the purification of natural gas or to produce synthesis gas, ammonia, alcohols and synthetic fuels. However, capturing CO_2 was not part of the process, unless it was necessary for the process. With the current transition with the CO_2 certificates on the free market, it provides the reduced emissions an economic value. Companies are also supported and rewarded at government level if they reduce their CO_2 emissions, where negative emissions achieve a higher value in sales than the emissions reduction. For these reasons CO_2 capture can generate added value both ecologically and economically. Depending on the location of the technical CO_2 capture in the process, it can be divided into pre- and post-combustion capture.

2.1 Pre-combustion capture

In the pre-combustion capture process, the CO_2 capture takes place before the energy source is combusted. This energy carrier is a hydrogen-rich gas, which is produced by an upstream gasification process. The main components of the gas are H_2, CO and CO_2. Furthermore, this gas is contaminated with accompanying substances. In order for the synthesis gas to be used, it must be processed accordingly. In the CO shift reaction, the CO is converted together with water vapor into H_2 and CO_2. After these process steps the CO_2 can then be separated by physical or chemical absorption processes. The remaining hydrogen is then used to generate electricity and heat.

2.2 Post-combustion capture

In post-combustion capture, the CO_2 is captured at the end of the combustion process. This has the advantage that existing plants can be retrofitted. For the CO_2 capture to take place, the flue gas must be dust-filtered, desulfurized, and denitrified. Depending on the type of capture, drying is also necessary. The possible post-combustion capture processes are absorption, the adsorption and the membrane process. These are shown in Figure 1.

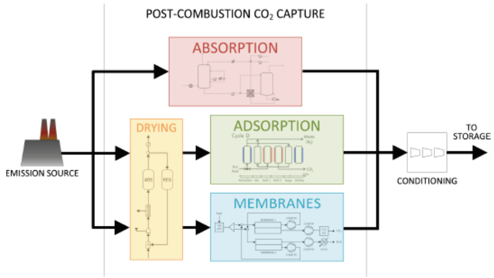

Figure 1: Post-combustion CO_2 capture, extracted from (Zanco et al., 2021).

2.2.1 Adsorption

One possibility for the CO_2 capture is adsorption. Adsorption is a thermal separation process, which runs discontinuously. The CO_2 that occurs in the flue gas stream adheres to the adsorbent. This binding can be chemical (chemisorption) or physical (physisorption). When the absorbent reaches saturation, it must be desorbed. This is done by a change in temperature (temperature-swing adsorption, TSA) or pressure (pressure-swing adsorption, PSA). Figure 2 shows the schematics of a TSA adsorber. In order to achieve a quasi-continuous process, at least a second adsorber is required, which works in interaction with the first (Reid, 2020). A possible cycle scheme, as shown in Figure 2. In TSA, the adsorber desorbs in a high temperature, as soon as it is saturated with the CO_2 from the flue gas. The advantage over desorption with the aid of a higher pressure lies on the one hand the possibility of using thermal energy at a relatively low-temperature level. This thermal energy is exergetically less valuable than the mechanical energy required to build up the pressure. Furthermore, unused heat flows from the combustion process can be used. Depending on the solvent used, the flue gas stream must be dewatered and cleaned beforehand.

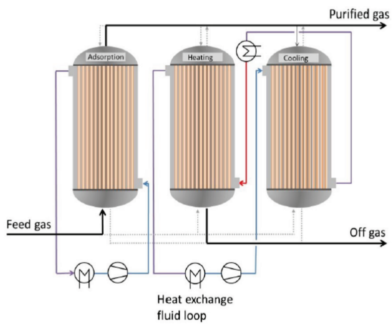

Figure 2: TSA schematics with tube bundle adsorber (Ried, 2020).

2.2.2 Membrane process

In the membrane process, the membrane material selection is important as it has different permeability with different gases. Thus, depending on the membrane material, a certain gas will penetrate the membrane better than the other gases. In this way, a gradual enrichment of CO_2 can take place before it is separated. A process schematic of the membrane process is shown in the Figure 3. The cost of capturing CO_2 by the membrane process is relatively low since the membrane materials are inexpensive. In order not to reduce the capture efficiency, it is necessary for the membrane process, as is the case with the adsorption process, to be preceded by flue gas dewatering and treatment.

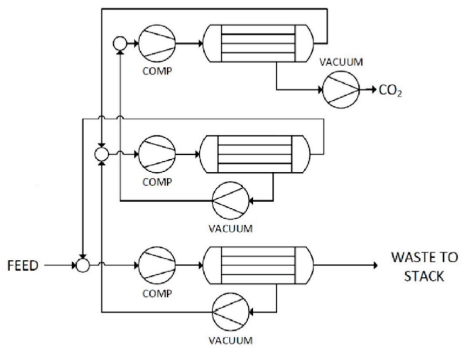

Figure 3: Schematics of the membrane adsorption process (Zanco et al., 2021).

2.2.3 Absorption

As the focus of the deliverable is on absorption, more focus is put on absorption and especially on the chemical absorption.

2.3 Types of absorption

In the absorption process, a large part of the total energy required for CO_2 capture is required for desorption. This required thermal energy depends on the scrubbing agent, the type of absorption, and the necessary purities of the CO_2 stream. In the case of physical absorption, a pressure difference between the absorber and desorber is often exploited in order to produce the desired gas component in a more energy-efficient way. The type of absorption is divided into physical, chemical, and hybrid absorption.

2.3.1 Chemical absorption

The main process of absorption consists of two columns. The flue gas to be cleaned enters the absorber column above the sump. There, the CO_2 of the flue gas stream is absorbed through chemical absorption by the aqueous scrubbing agent, which is introduced in the upper part of the absorber. At the top of the absorber, the low-CO_2 gas stream is then discharged. The higher the purity of this low-CO_2 gas stream, the more stages/packing height the absorber must have. This increases the cost. The CO_2-rich is then fed to the desorber and let into its upper section. In the desorber, CO_2-enriched solvent is heated so that the CO_2 can be expelled more effectively. This expulsion of CO_2 from the solvent is achieved by the rising water vapor, which is evaporated in the lower section of the desorber. The head condenser separates the condensing water and the gaseous CO_2. The gaseous CO_2 is now in the system and can be further treated. The water is returned to the desorber. The regenerated and low-CO_2 scrubbing agent is discharged into the sump of the desorber before it is cooled and fed to the absorber. To ensure the cooling of the low-CO_2 before entering the absorber and the heating of the high heating of the high-CO_2 scrubbing agent in the desorber more energy-efficiently, heat transfer takes place between these mass flows. In the downcomer and desorber, mass and heat transfers are improved by using internal structures such as trays and packings. This improves the separation efficiency (Ohle, 2009). In Figure 4, it can be seen that there are different heating and cooling requirements. The aim is to integrate this process optimally, for example, using its excess waste heat. In chemical absorption, the gas component is separated and the absorber forms a loose- or solid-chemical bond. Chemical scrubbing agents are used in a particular way, where a small amount of the gas component has to be separated at low partial pressure. Since the chemical scrubbing agents act selectively with the various gas components, high purities can be achieved. Some examples of chemical washing agents are Monoethanolamine (MEA), Diethanolamine (DEA), or Methyldiethanolamine (MDEA). These chemical solvent are particularly suitable for the post-combustion process with a low partial pressure.

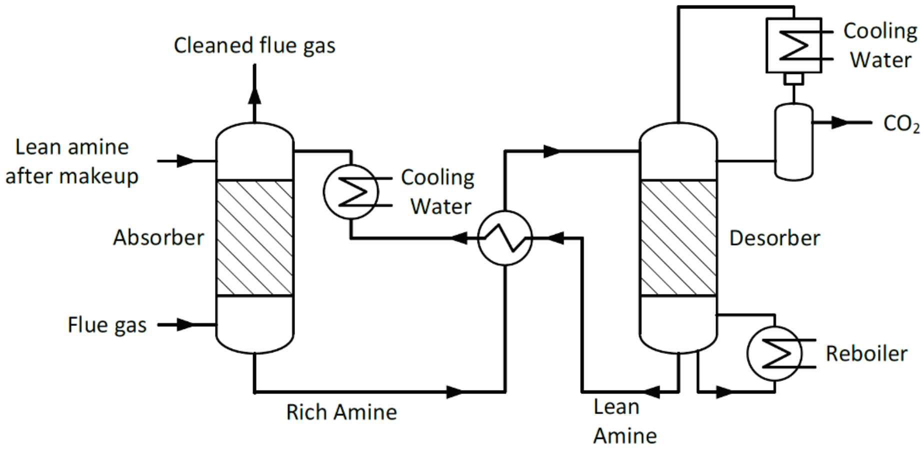

Figure 4: Schematic of a standard CO_2 absorption process.

2.3.2 Selection of solvents

Particularly in the case of hybrid agents, many new solvents have emerged. These are expected to offer large energy savings in CO_2 capture. The chemical solvent, MEA, is still the most widely used scrubbing agent for capture processes based on aqueous alkanolamines. The reason for this is the high absorption rate and the low solvent costs. The thermal energy demand due to the high reaction temperatures with CO_2 (85 kJ/mol CO_2) is high (Schäffer, 2013). An advantage of the DEA chemical solvent is the lower thermal energy demand due to the reaction with CO_2 (79 kJ/mol CO_2) (Schäffer, 2013). However, since corrosive decomposition products at high CO_2 concentrations are a concern, it is not an option for the present work. Manufacturers such as Vattenfall or the SINTEF research group are working on alternative solvents with improved properties (Kvamsdal et al., 2011). Energy efficiency and operating costs could be improved in the future.

MEA is the most tested and researched solvent available for this application in the simulation software CHEMCAD, which is used for this work. This solvent is implemented and presented in a reference example. It is therefore to use the MEA solvent for this work.

2.4 Process embedding

The focus of this work is on the consideration of the absorption plant. Only this part is modeled in detail in the simulation software CHEMCAD. It is important to understand how the flue gas stream has to be pretreated and the CO_2 mass flow must be post-treated. The pretreatment and post-treatment are considered with data from literature.

2.4.1 Flue gas pretreatment

Amine scrubbing works well with clean flue gas. However, if the flue gas, is contaminated, the efficiency of CO_2 decreases, since nitrogen oxides- and sulfur-containing gas components also bind with the scrubbing agent. Depending on the nitrogen oxide and sulfur contents, the flue gas stream has to be desulfurized and purified of nitrogen oxides. A further process stages are drying and cooling of the flue gas stream.

2.4.2 Process configurations

Different process configurations can reduce the energy consumption of the deposition process. Here, a trade-off has to be found between energy savings and additional investment and operating costs. The findings obtained from the work of (Xue et al., 2017) serve as a reference point for possible process optimizations.

2.4.3 Post-treatment CO_2 stream

As a post-treatment step, the gaseous CO_2 must be compressed and liquefied, so that it can be transported. This investment and operating costs are taken into account with estimates from literature.

2.5 Reference CO_2 capture plants

The simulation model, which represents the absorption process must be created with credible data. Based on comparable reference plants and empirical values from literature, sound assumptions can be made. With particular interest on the CO_2 absorption plants from Esbjerg and Brindisi, as they have similar CO_2 capture rates (Kvamsdal et al., 2011), (Flø et al., 2016). A CO_2 absorption plant, which is planned in Switzerland is expected to be built at the waste incineration plant.

3 Case study

This work studies the CO_2 capture by absorption at a wood-fired power plant. The considered CHP is planned and is to be operated by the combustion of regionally delivered wood-waste and forest chips to produce CO_2-neutral electricity and heat. The heat will be used at two different temperature and pressure levels. The process heat will be supplied to the customers in the vicinity of the CHP. A new district heating network is to be built for the use of district heat.

3.1 Actual state of the CHP plant

The combustion of wood waste and forest chips takes place with a capacity of 16.9 MW boiler. The system with grate firing enables staged combustion. The combustion temperature is controlled by air preheating, air volume control, and flue gas recirculation in such a way that the pollutant components such as CO, and thermal NOx formation can be minimized (Hemmerlein, 2022). In order to further reduce the NOx reduction, a urea solution is injected (SNCR process). At this point, the flue gases have a temperature of approx. 150°C. Their further treatment is through an electrostatic precipitator and a fabric filter. Upstream of the stack, the flue gas is analyzed to check the efficiency of the flue gas purification system and compliance with the Clean Air Ordinance (LRV). The pressure of the upstream of the steam turbine is at a temperature of 425 °C and 65 bar(a). In the steam turbine, the pressure is reduced to 0.1 bar(a). The mechanical power generated in the steam turbine is fed by a 4.5 MW synchronous generator into the 20 kV grid of the local power plant. In addition to the two pressure and temperature levels, steam decoupling takes place. For the process heat, this decoupling takes place at 6.95 bar(a) and 188 °C. The district heat is decoupled at 1.2 bar(a) and 105 °C. These data can be seen in Figure 5. If there is an increased demand for thermal energy, power production can be throttled. This makes it possible to meet peak loads without the additional use of fossil fuels. Depending on the ambient temperatures, a different demand for process and especially for district heating, recooling is necessary for summer. A part of the thermal energy thus remains unused.

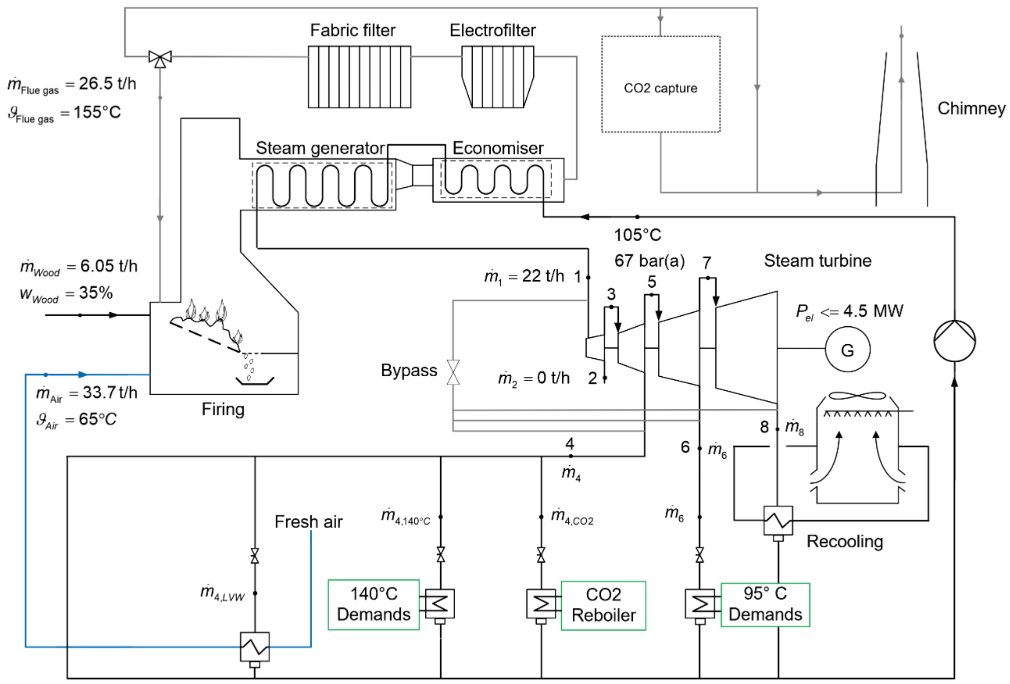

Figure 5: Overall flowsheet of the case study

3.2 Extension of the CHP plant with CO2 capture

A CHP plant is already CO_2-neutral in normal operation since the CO_2 emitted after combustion is recaptured by renewable. It is also possible to capture the CO_2 from the flue gas. This could also be classified as negative emissions. The possible processes are described in chapter 2.2. The next section presents the integration post-combustion capture with a CO_2 absorption plant. Criteria for the design and evaluation are the technical feasibility and energetic, economic, and ecological changes.

If the CHP plant is expanded with a CO_2 capture system, it is advisable to replace the reboiler of the CO_2 absorption plant by using the thermal energy on the process heat stage. If steam is used in the process heat stage for the CO_2 capture, electricity production is reduced. This leads to a decrease in the revenue from electricity sales and must be taken into account. The operation of other components of the CO_2 absorption plant also requires thermal and electrical energy. These additional costs reduce the efficiency of the CHP plant. From the work of (Ohle, 2009), it shows that the use of a solvent with a 30% MEA content can result in an efficiency reduction of up to 14%. This includes the capture as well as the CO_2 liquefaction. Part of this work is to optimize the design of the CO_2 absorption plant to keep this reduction in efficiency as small as possible.

3.3 Consideration of process and district heating network

At an ambient temperature of around 20°C, a large proportion of the thermal energy remains unused. The reason is the high thermal power demand of the district heating network at cold ambient temperatures in winter temperatures and the lower demand at warm temperatures in summer when the heating periods are over. Also, in the transitional period and especially in summer, this unused thermal energy is available for CO_2 capture. Since the district heating network has not yet been built and the exact demand is not yet known, scenarios are developed for the heat demand. For this purpose, a real heating curve of a comparable district heating network is generated. Additionally, the annual temperature curve and the maximum available thermal district heating capacity of the CHP are considered. For the process heat network, data are available from the heat consumers. With this, the annual course of the CHP can be calculated. This makes it clear which thermal outputs are unused and can be used for the operation of the CO_2 absorption plant. This compilation does not show the real demand, but a sufficient basis for the possible fluctuations.

3.4 Analysis of available services

Figure 6 shows the composition of the total capacity of the considered CHP. These powers are electrical power, thermal energy for air preheating, thermal energy for air preheating, process heat, the energy of district heat, and as well energy losses. In the previous design, a large amount of thermal energy remains unused in summer. Figure 6 shows the thermal energy that can be used for the operation of the CO_2 absorption plant. This is composed of the hourly values, which are summed up. In this way, the annual energy for the reboiler can be calculated. Using this graph, it is possible to estimate how many hours per year of thermal power is available for the operation of the CO_2 absorption plant. Since the CO_2 absorption plant cannot operate at any given partial load, the thermal capacity for which the CO_2 absorption plant has to be designed accordingly. In chapter 6.2, several simulations for different boundary conditions are created and compared. Based on the annual hydrograph, the available power is then calculated so that a rough estimate for the annual operating data can be made.

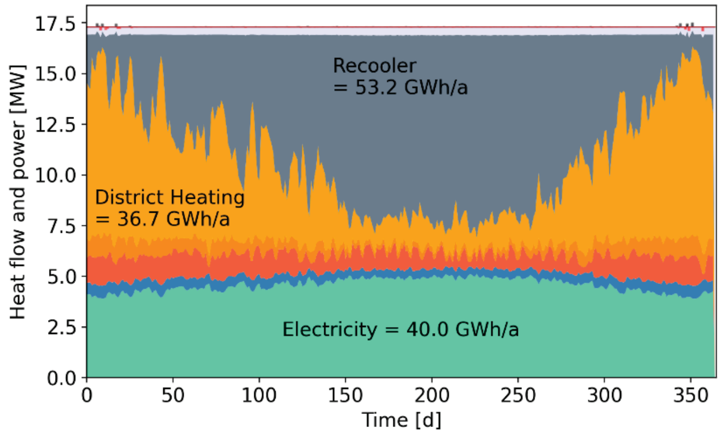

Figure 6: Representation of CHP outputs and annual thermal energy of the reboiler at 100% flue gas treatment.

3.5 Conclusion available power reboiler

From the consideration of the annual hydrographs, it is reasonable to consider the design of CO_2 absorption plants for different reboiler capacities. The variation of the required capacity is achieved by treating only a part of the flue gas volume. In a second step, it will then be tested how this design runs with various reboiler capacities. This work under partial load is important, as the available thermal energy varies greatly. The percentage treatment of the flue gas quantities which are considered are defined in chapter 5.2 and are 100%, 65%, and 30%. This results in different expansion stages in the base case, as well as a large and a small CO_2 absorption plant size. The small CO_2 absorption plant has a smaller reboiler capacity and captures less and separates less CO_2, but the running times are longer. The large CO_2 absorption plant separates more CO_2 but achieves fewer full load hours. The goal is to roughly calculate based on the simulation data which plant size is the most suitable.

3.6 Characterization of input and output variables

The design the CO_2 absorption plant in the software CHEMCAD software, it is necessary to make to carry out rough estimations regarding the design on simplified manual calculations. These data then serve as input parameters and for comparison and plausibility of the data from the simulation. The CHP is used at different ambient temperatures and load cases. For a first estimation of the conditions and the system modeling, the operating case full load at an ambient temperature of 20°C and a wood moisture content of 35% is considered. The input currents that occur during the absorption process is extracted from VAS AG (2020). In the case of a capture process with the MEA scrubbing agent, CO_2 capture rates of 95% can be achieved (Schäffer, 2013). In the CO_2 absorption plant design considers which operating parameter and which component design suitable for the treatment of this flue gas are suitable.

3.6.1 Flue gas stream

The exact flue gas composition varies depending on the wood moisture content and the excess air number. For the load case presented in the previous section, the flue gas composition is as follows:

Table 1: Gas composition of the flue gas.

Flue gas composition	Mass flow rate [kg/h]	Mass %
CO_2	7'143	19.1
H_2O	4'196	11.2
O_2	2'078	5.6
N_2	23'926	64.1

The outlet temperature of the flue gas stream is 150 °C. These data form the input values for the simulation in the CHEMCAD software. Lower pollutants in the flue gas, through combustion or better pretreatment of the flue gas stream, the lower the degeneration of the scrubbing agent. The emission values are taken from the data VAS AG, which designed the CHP plant. These pollutant emissions impair the operation of the CO_2 absorption plant. With a more detailed design, it is necessary to check whether an additional flue gas treatment is necessary.

3.6.2 Results of balancing

For the balancing of the mass and energy flows, different assumptions have to be made. The validation of the simulation is based on data from the literature. Since the scrubbing agent MEA, which is already often used, the confidence in the simulation results with a suitable validation is high.

4 Simulation of the CO_2 absorption plant

This chapter defines the implementation of the CO_2 absorption process into the CHEMCAD program. This includes the component selection, the determination of the thermodynamic model, and the definition of the boundary conditions. Three CO_2 absorption plant sizes are compared, and different process optimizations are presented. The validation of the simulation model and the presentation of the multistage process simulation round off this chapter. Figure 7 shows the basic simulation design of the absorption plant. The most important components are the water separator, the absorption column, the heat exchanger, and the desorber. In chapter 5, their operating behavior is a function of various parameters in order to improve the absorption process.

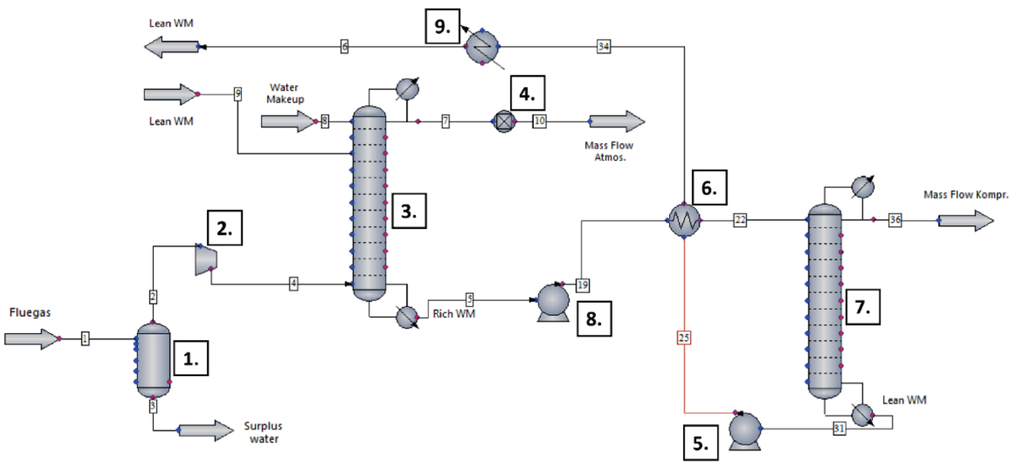

Figure 7: CHEMCAD simulation of CO_2 absorption through amine-washing.

4.1 Modeling in CHEMCAD

The CO_2 capture into a scrubbing solution is a reactive absorption/desorption process. In the software CHEMCAD, the amine model is used for amine solutions. The phenomena that occur in a reactive absorption/desorption process, are described in the book CO_2 Capture by Reactive Absorption Stripping Modeling, Analysis and Design (Madeddu et al., 2019).

The following five issues affect this process:

- Components and thermodynamics
- Chemical reactions
- Material and energy balances
- Simultaneous heat and mass transfer
- Fluid dynamics

4.1.1 Components and model selection

In the first part of the process in the absorber, two phases are involved. On the one hand, the incoming gaseous is composed of nitrogen (N), oxygen (O), water vapor (H_2O), and carbon dioxide (CO_2). The liquid phase flows into the head of the absorber. Its components are CO_2, H_2O, and MEA. After the components have been selected, different modeling approaches for the phase/reaction equations and the enthalpy models can be chosen (Pereira & Vega, 2018). In the CHEMCAD program, for the CO_2 capture with amine solvents the K-model "Amines" is provided. This is based on the Kent- Eisenberg (1976) model, which is a semi-empirical model. The Kent-Eisenberg model neglects the non-ideal processes in the vapor and liquid phase and the correctness of the equilibrium values is achieved by a fit using experimental data.

4.1.2 Vapor-Liquid-Equilibrium-Model

During the simulation of an absorber or desorber, the vapor-liquid equilibrium (VLE) approach is selected, and a vapor-liquid equilibrium is calculated at each separation stage. This setting is mostly used by default. This model is used first, to analyze the basic behavior of the system. In a further step the change to the mass transfer model. The reasons are the VLE model, as well as the simpler application.

4.1.3 Packed Column Mass Transfer Model

The packed column mass transfer and the tray column mass transfer use the Maxwell-Stefan-diffusivities as well as empirical data to generate a matrix of the general mass transfer coefficients. Additionally, in the CHEMCAD software, the most important column properties can be entered in a dialog window. With these inputs, a more precise simulation of the process and a component design can take place. In chapter 6.1 there is a comparison between the VLE and the mass-transfer model. For the following calculations, the results of the mass transfer model are used.

4.2 Simulation of the entire system

For the CO_2 absorption system to be designed, it must function as an overall system. However, when simulating the entire system, there is a greater chance of encountering convergence problems. The structure of the overall simulation is based on the reference simulation for amine solvents from CHEMCAD. This is adapted to the operating conditions. After the presentation of the mode of operation and the values of the individual components and the explanation of the entire functional sequence, a parameter study is developed. This provides information about possible operating and delivery parameters and which design of the CO_2 absorption system works best.

4.2.1 Boundary conditions

Possible boundary conditions for the overall simulation are estimates from the partial simulations, validated by reference plants with the same scrubbing MEA (30% part by weight). The loading of the solvent is usually given in the literature CO_2/mol MEA in the literature. Therefore, this specification is also used here.

Table 2: Possible boundary conditions for the CHEMCAD simulation.

Parameter	Range of Values	Units	References
Loading Solvent Rich	0.47 – 0.52	Mol CO2/mol MEA	0.5 – 0.6 (Madeddu et al., 2019)
Loading Solvent Lean	0.185 – 0.37	Mol CO2/mol MEA	0.2 – 0.4 (Madeddu et al., 2019)
Separation rate	75 – 95	%	Up to 95% (Schäffer, 2013)
Thermal energy Reboiler	3.4	GJ/t CO2	3.1 – 3.7 (Knudsen et al. 2009)
Temperature Solvent Rich	90 – 110	°C	Up to 120 (Schäffer, 2013)
Temperature Fluegas	30 – 60	°C	Ca. 45 (Schäffer, 2013)

4.2.2 Structure of the overall simulation

The input parameters for the flue gas flow are known and stored in CHEMCAD. For the thermophysical settings, the amine model is used. The components numbered in Figure 7 and their basic parameters will be basic parameters are discussed below.

1. Flash
The flash component cools the flue gas stream to a predefined temperature. When more cooling is needed, more cooling water is required. In particular as the water in the flue gas condenses above a temperature of approx. 50 °C, causing condensation in the flue gas, releasing thermal energy which must be additionally dissipated.

2. Blower
To ensure that the flue gas entering the absorber is at the required pressure level, the pressure must be increased to a predefined value. This is done in the blower. In addition, the temperature of the flue gas rises slightly due to the polytropic of the flue gas increasing slightly.

3. Absorber
In the absorption column, the flue gas is cleaned, depending on temperature and pressure, as well as the required purity. A different amount of scrubbing agent is used, and the dimensioning is different.

4. Controller
Through the controller component, for example, input stream data can be varied until a parameter of a mass flow reaches a predefined value. The feed-backward controller used varies the mass flow of the incoming scrubbing agent lean until the mass flow of CO_2 in the exhaust gas reaches a defined value.

5. Solvent rich pump
In the case of the solvent rich pump, the pressure of the fluid is increased to a value necessary for the absorption process.

6. Heat exchanger solvent lean/rich
The solvent rich exit the absorber at a low temperature. For the inlet in the desorber, however, a high temperature is more suitable. In addition, the inlet temperature of the absorber should be as low as

possible. It is therefore energetically to install a heat exchanger between the stock streams of rich solvent and lean solvent, to adapt their temperatures to the process (heat recovery).

7 Desorber
In the desorber, the CO_2 is driven out of the solvent and separated. The optimum operating conditions are low pressures and high temperatures. To drive the CO_2 from the solvent, a reboiler is installed in the lower part of the desorber, which depends on the capacity and temperature level of the solvent. Furthermore, at the parameters of the desorber, the condensation temperature is defined. As with the absorber Depending on the operating conditions, different reboiler and desorber dimensions are required to achieve the desired properties of the CO_2 stream and of the solvent composition.

8. Pump solvent lean
The pump for the mass flow of solvent lean increases the pressure to a value that is necessary for the absorption process.

9. Lean solvent cooler
In most cases, a greater temperature change is required for the Lean solvent than for the rich solvent. For this reason, an additional heat exchanger is installed. Depending on the mass flow of the lean solvent and its required temperature level there is a different cooling water requirement.

4.3 Process optimizations

The absorption process can be improved by simple process modifications. These adjustments are selected and evaluated based on various studies. It is clear that in the more detailed design, further process optimizations will be made. By evaluating the most important opportunities for improvement, however, it is already possible to better estimate what a possible final solution will look like. The work by (Xue et al., 2017) compares different CO_2 absorption plant configurations for their energy requirements. Possible improvements can be derived and used for validation of the obtained data used.

4.3.1 Variant 1: Rich solvent split (RSS)

The process variant RSS requires only one additional component, a splitter for the rich solvent mass flow. This is then installed upstream of the rich/lean heat exchanger (number 6. in Figure 7). Part of the mass flow thus reaches a higher inlet temperature, while the other part remains at a lower temperature level. The colder mass flow is admitted to the upper part of the desorber, while the warmer part enters the desorber at a lower level. This leads to energy savings in the desorption process.

4.3.2 Variant 2: Heat recovery of flue gas

In this process optimization, the scrubber rich stream is further heated by the flue gas flow before entering the desorber. This brings two positive effects, the desorption process becomes more energy-efficient and results in less cooling water needed for flash cooling, since the flue gas stream already enters cooler.

4.3.3 Variant 3: Intercooling

Absorption is better when low temperatures prevail. The chemical changes in the flue gas and the scrubbing agent, heat is released, which increases the temperatures in the absorber. This decreases the separation efficiency. If now the scrubbing agent and gas in the absorber are split out at a certain level, it is cooled and reintroduced at a certain level, and the separation efficiency increases.

4.4 Plant variables of the absorption process

As discussed in Section 3.5, different CO_2 absorption plant designs for different reboiler capacities are considered. It is therefore useful to create and analyze possible simulation models. The following flue gas flows, separation rates, and reboiler capacities are examined:

Table 3: Value range of important boundary conditions in the different simulation models.

Parameter	Range of Values	Unit
Flue gas mass flow	11'203 – 37'342	kg/h
Separation rate	20 – 98	%
Process heat	1.5	MW
District heat	2 – 10	MW
Reboiler power	0.5 – 8	MW

4.5 Validation of the simulation model

In order to validate the operating behaviour of the CO_2 absorption plant, reasonable operating parameters, expected energy costs, and possible CO_2 savings can be determined the model must be validated. Otherwise, there is a risk that incorrect data will be generated, leading to lead to inaccurate interpretations. It should be mentioned that the use of the VLE described in section 4.1.2 allows for certain inaccuracies. The basic behaviour with changing parameters remains the same and thus shows well which design conditions are reasonable. Since it is difficult to obtain operating data of a CO_2 absorption plant in the same order of magnitude and to obtain and to obtain the corresponding operating data, the validation refers to different works and experimental data. In the following the comparison between data obtained from the simulation with reference work.

4.5.1 Comparison 1: required reboiler energy per captured kg CO_2

A validation approach made by Gervasi et al. (2014) is the comparison with the required thermal reboiler energy per kg of CO_2 captured. This parameter is particularly interesting since a lot of thermal energy at a high-temperature level is an important evaluation factor. Typical values for 90% CO_2 capture and a mass fraction of 30% MEA in the solvent are 3.7 - 4 GJ/t CO_2. (Kvamsdal et al., 2011), (Gervasi et al., 2014), (Pasini et al., 2011).

4.5.2 Comparison 2: Solvent loadings lean/rich

A further comparison, which can be validated based on experience and literature values, is possible solvent loads. Typical values are solvent lean 0.2 mol CO_2/mol MEA and with the rich solvent 0.5 mol CO_2/mol MEA (Madeddu et al., 2019), (Li et al., 2016), (Haar et al., 2017). These values are in agreement with the simulation data.

4.5.3 Comparison 3: Temperature curves absorber and desorber

Since thermal energy is released during chemical absorption, the temperature changes during the absorption and desorption processes. The temperature, which changes as a function of the absorber /desorber stage, provides a further reference value for plant validation. The temperature behavior of the absorber and desorber columns is compared with the data of (Cousins et al., 2012). The experimental data from a real CO_2 absorption plant, as well as simulated data. The real data refer to the Tarong CO_2 absorption plant. The flue gas volume and the captured CO_2 are lower, but the flue gas composition is comparable to the CO_2 absorption plant. Since various test runs were carried out in the operating data, various test runs were carried out, the data of the test runs are used for the validation which, as far as possible, corresponds to the operating data of the CO_2 absorption plant to be developed. The differences in the data comparison can be caused due to different parameters.

4.5.4 Comparison 4: Loadings in the absorber

Finally, the data of the loadings in the absorber are compared with experimental data. These data originate from the same work as in comparison three (Cousins et al., 2012).

4.5.5 Conclusions

Based on the four different comparison parameters from different works and experimentally determined data and their agreement, the simulation model developed simulation model seems plausible. It can be used for parameter studies and other evaluations.

5 Parameter studies of the VLE model

The parameter study for the determination of the appropriate operating parameters is based on the multistage section explained, where the VLE model is used to design an optimal CO_2 absorption plant. Also, the creation of different plant configurations and the different CO_2 absorption plant sizes can be evaluated. Before the variable parameters study, CO_2 absorption plant configurations, district heating scenarios, and evaluation data are evaluated. In the second step, the parameter study takes place based on the adjusted operating parameters and further evaluations can be made. Various programs are used to carry out the parameter study. First, the simulation of the data to is adapted accordingly. If the simulation converges, the data are stored in an Excel file. This contains approximately 80 values, which can be called up as required. The required data is then read into a Python program data, calculated, and displayed graphically. The interpretation of the data takes place in the simulation. The most important findings from the parameter study are discussed in this chapter. The parameter study is used to verify three main aspects. Firstly, it is necessary to check which operating parameters are optimally suited. This concerns the results such as the CO2 capture rate, loads of solvent required make-up MEA and water, the temperature level for flue gas cooling, and other values. Once these parameters have been determined by the first simulation, in the second step the different CO_2 absorption plant variables are compared and discussed.

5.1 Evaluation of operating parameters

The program CHEMCAD offers an optimization function, which allows a multiparameter study. With this procedure, all parameter combinations in a predefined range are compared. The optimum related to the simulation is recognized. The stepwise analysis of the single parameters, as applied here, neglects possible interactions between parameters. It is therefore possible that the global optimum is not achieved. The detailed optimization is too complex and costly within the scope of this work. The most important evaluation factors are the thermal energy required to capture one ton of CO_2, the required cooling water flows, the electrical energy required to operate the CO_2 absorption plant, and the required make-up MEA. The following operating parameters allow an optimized CO_2 absorption plant- operation and will be used from now on. When switching to the packed column mass transfer model, changes may occur, which will be discussed accordingly.

Table 4: New operating parameters which are defined by parameter study.

Parameter	New Value	Unit
Temperature after flash	45	°C
Blower pressure	1.1	Bar
Absorber stages	16	-
Pump lean	1.5	Bar
Pump rich	3	Bar
Temperature solvent before desorber	110	°C
Desorber stages	18	-
Desorber feed stage	4	-

5.2 Evaluation on plant size

Chapter 3.4 shows which thermal energy remains unused and to what extent it can be used for the operation of the reboiler. This unused thermal energy can be utilized by different-sized CO_2 absorption systems. Depending on the CO_2 absorption plant size, only part of the flue gas flow is treated. To determine which flue gas flows, three reboiler capacities are defined, at which the CO_2 absorption plants achieve 90% CO2 capture to the incoming flue gas volume. The three CO_2 absorption plant sizes selected are 100%, 65%, and 30% flue gas treatment. Due to the fixed maximum capacities, the hourly outputs of the overall system as well as the resulting annual cycles can be processed.

5.2.1 100% Flue gas treatment

Figure 8 shows the changing power demand of the different heat utilizations of the CHP. The light red marked area shows this thermal energy. Here a maximum thermal power of 6.6 MW is used. It can be seen that in summer a part of the thermal energy has to be recooled (dark gray area). It is not possible to use all of the thermal energy. On the one hand, at warm ambient temperatures too much thermal energy is available, and, on the other hand, too little thermal energy is available at cold temperatures to operate the CO_2 absorption system. It is to be checked with which reboiler capacities the 100% flue gas treatment still works well and which different CO_2 capture rates, mass flows, and CO_2 absorption plant data for these different thermal capacities are achieved.

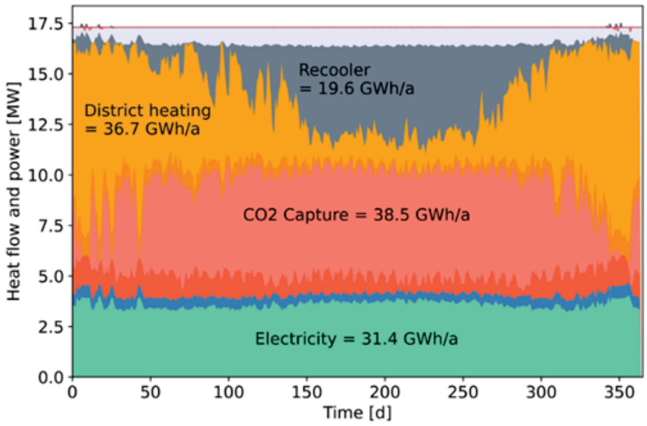

Figure 8: Representation of CHP outputs at 100% flue gas treatment.

5.2.2 65% Flue gas treatment

If the CO_2 absorption plant is operated at a maximum of 4 MW reboiler capacity and 65% of the flue gas is treated, a larger amount of thermal energy remains unused, which has to be recooled. For this purpose, the CO_2 absorption plant can be operated for a longer time at lower available thermal powers. The exact analysis regarding the power range, in which the CO_2 absorption plant works well, is also important.

5.2.3 30% flue gas treatment

In the case of 2 MW, higher thermal energy is recooled. Therefore, it is expected that the CO_2 absorption plant has even more operating hours. By recording the annual operating data this CO_2 absorption plant can also be compared with the others.

5.2.4 Comparison of plant sizes

As explained in the above section, the CO_2 absorption plants treat different amount of flue gas depending on their size. In the 6.6 MW case, 100%, the 4 MW case 65%, and the 2 MW case 30% of the flue gas is treated. For the determination of the most suitable CO_2 absorption plant size and the differences between the individual variants, a higher-level comparison is useful. This makes it possible to estimate which solution offers the most potential. Some parameters are more or less the same for all operating cases. This concerns the values of the purity of the CO_2 stream for compression and loads of the solvent. Therefore, these data are not further compared.

5.2.5 Percentage deposition

One of the most important parameters is the percentage of the total emitted CO_2 of the CHP plant. Figure 9 shows the percentages of capture rates, which vary depending on the flue gas entering the CO_2 absorption plant and the existing reboiler capacities. With the same reboiler capacity but a different flue gas volume to be treated, the same amount of CO_2 is captured. In addition, 30% or 65% treatment, a maximum of this amount of the total CO_2 can be captured. This point is reached when the curve flattens out and, despite an increasing reboiler output, no more CO_2 can be captured. With Figure 9, it is easy to estimate which available thermal reboiler capacities and what amount of CO_2 can be captured. Since these simulations are based on the VLE model, it is possible that with a changeover to the packed column mass transfer model, the possible operating range may be reduced. However, for a preliminary comparison of the different CO_2 absorption plant sizes, these data are sufficient.

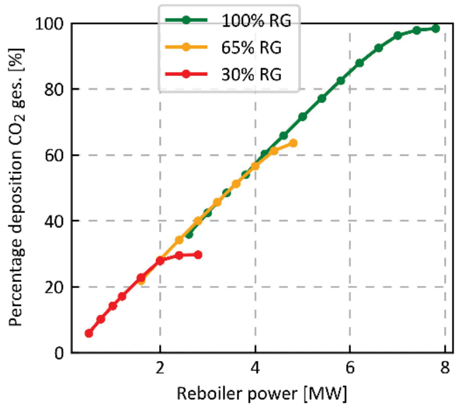

Figure 9: Comparison of process separation for different plant sizes with boundary conditions: Close Loop Model, VLE model, without process optimization.

6 Parameter study mass transfer model

The simulations and evaluations carried out so far show the behavior with the use of the VLE model. The mass-transfer model also considers the kinetics in the absorber and desorber columns. The obtained more accurately represent reality and are more suitable for the comparison of the different CO_2 absorption plant designs. The influence of the packing selection on the partial load behavior is an important aspect to be examined. It is important to find a suitable packing, which covers a large operating range and allows an efficient CO_2 absorption plant operation.

6.1 Overall comparison of plant size and VLE/mass transfer model

By directly comparing the capture rates as a function of the CO_2 absorption plant size and the reboiler capacity in Figure 10a, it can be seen that when the CO_2 absorption plant is operated in the appropriate operating range, only the reboiler capacity influences the amount of CO_2 captured. However, the size of the CO_2 absorption plant does not. At the consideration of the reboiler capacity of for example 4 MW, the CO_2 absorption plant which treats 100% of the flue gas, captures the same amount of CO_2, although it operates in a partial load mode. Since the large CO2 absorption plant processes more flue gas, it captures less in percentage terms. This makes the process more efficient. Since there is also a larger quantity of solvent, the difference in the loads must be smaller. This also makes desorption more efficient. The linear behavior offers to graphically determine which capture rates can be achieved at which power levels. This procedure is applied in section 6.2. Figure 10b shows the efficiency of the CO_2 capture as a function of the power range and the flue gas volume to be treated. The larger the maximum power and the amount of flue gas to be treated, the larger is also the power range in which the CO_2 absorption plant can be operated.

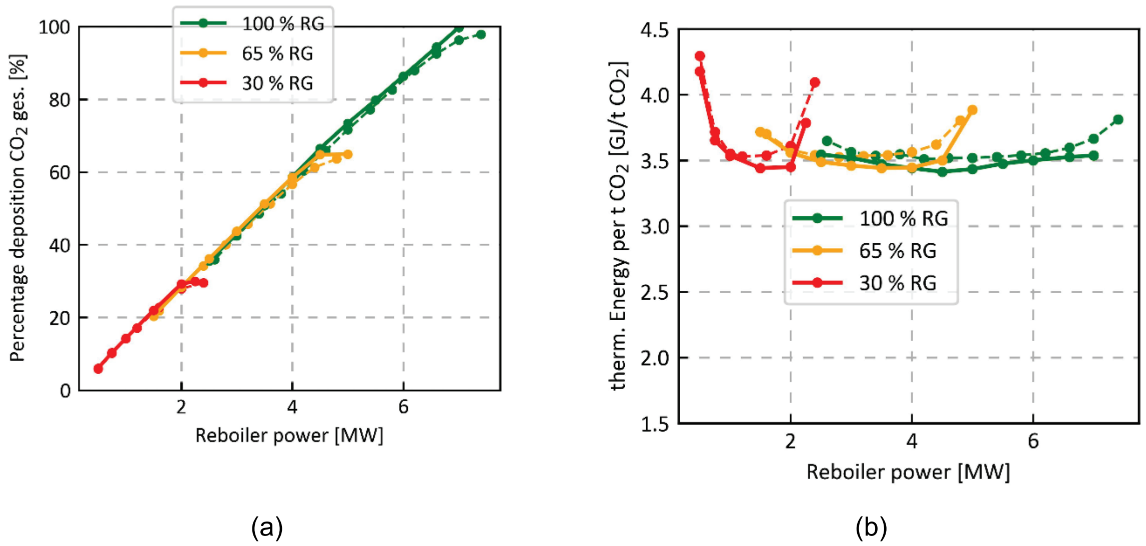

Figure 10: Comparison of VLE (dashed) and mass-transfer model (solid) with boundary conditions: close loop model, without process optimation for (a) potential of CO_2 capture, (b) and thermal energy needed per t of CO_2 capture.

6.2 Calculation of the annual operating data

The annual operating data are calculated by the available thermal power at the reboiler and by the simulation data. At an existing reboiler capacity, it is possible to see in the output Excel files of CHEMCAD, which operating conditions exist. It shows for example how much CO2 is captured and how much cooling water is required. With all the hourly performance values, the annual operating data can be summed up. This procedure is applied to the flue gas treatment of 100%, 65%, and 30%. The capacity range in which the corresponding CO_2 absorption plants operate is determined in the previous section. With the results from the calculations, a comparison of the possible solution variants.

6.2.1 Annual CO_2 captured

Figure 11 shows the annual amount of CO_2 that can be captured, depending on the amount of flue gas to be treated and the size of the CO_2 absorption plant size. The value is calculated by summing up all the hourly CO_2 mass flows leaving the desorber. It can be seen that the CO_2 absorption plant for 100 % treatment with approx. 5'700 h, operates the least under full load. This is because the high Reboiler power needed is only available at warm outside temperatures. The design sizes 65 % and 30 % achieve more full load hours. The total runtime does not vary very much. This is due to the fact that with a larger CO_2 absorption plant design, its capacity range is also larger and the CO_2 absorption plant reaches long annual runtimes. The annual CO_2 captured with the design for 100 % flue gas treatment is 48'300 tons. This corresponds to 77.3 % capture of the annual CO_2 emissions of the CHP. The CO_2 absorption plant for a 65 % treatment captures 34'200 tons of CO_2 per year, which is 54.7 % of the annual CO_2 emissions of the CHP plant. The treatment of 30 % of the flue gas results in an annual CO_2 capture of 17'000 tons. This corresponds to a percentage capture of the annual CO_2 emissions of the CHP of 27.2 %.

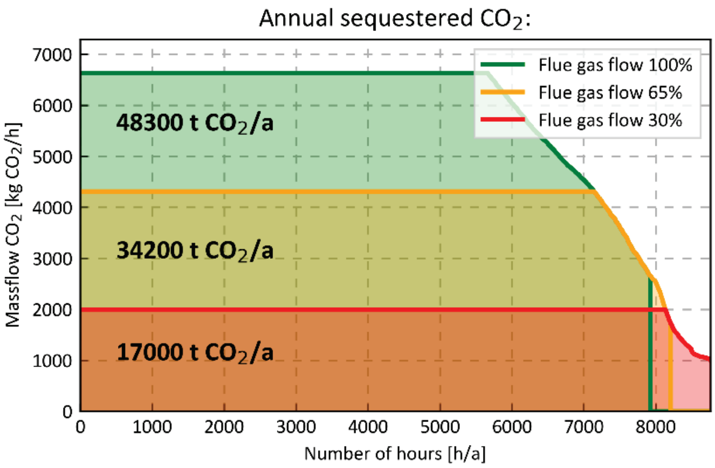

Figure 11: Annual CO_2 captured.

7 Process Integration

In this section, the considered section for pinch analysis are:

I. The heating demand and electricity production of the CHP plant

II. The structure and demand of the different heating networks.

III. The amine gas treating process.

IV. The integration the amine gas treating process into the CHP plant.

For the heating demands and electricity production are laid out as in Section 3.1 and the amine-gas treating process as in Section 4.

7.1 District heating network

The heating demand of the district heating network is estimated via the heating curve of the district heating network. Extraction steam at 1.2 bar(a) is used to supply heat to the heat distribution network. Further, the exhaust steam of the condensate tank is condensed in a further condenser to supply heat to the network. Three thermal energy storages (TES) with a volume of 46 m^3 each are used to smooth out peak loads. If the heating demand of the district heating networks exceeds the capacity of the extraction steam line, two oil fired emergency boilers with a heating capacity of 5 MW each can be used to compensate the heat deficit.

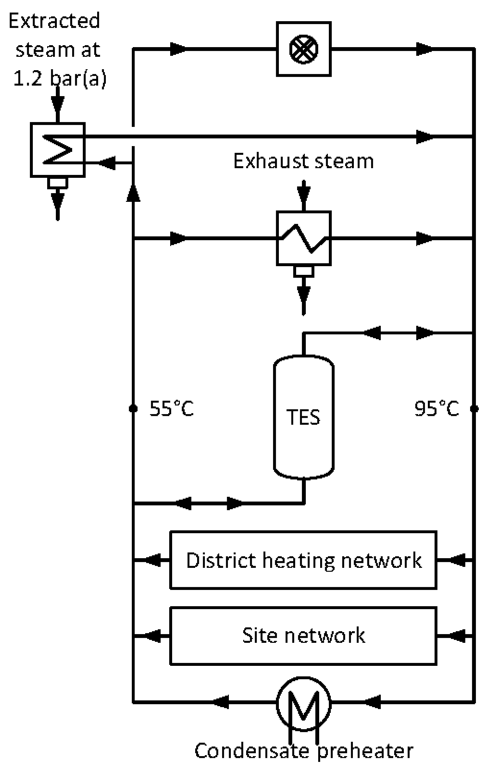

Figure 12: Detail view of the basic structure of the heat

The nominal supply temperature of the network is 95 °C. The installed 'main'-condenser for the heat distribution network has a nominal power of 10 MW and the exhaust steam condenser has a nominal power of 250 kW.

7.2 Process heat network

The process heating network has a maximal heating demand of 1.5 MW. The annual duration line is shown in Figure 13. The nominal supply temperature of the network is 145 °C with a nominal return temperature of 125 °C.

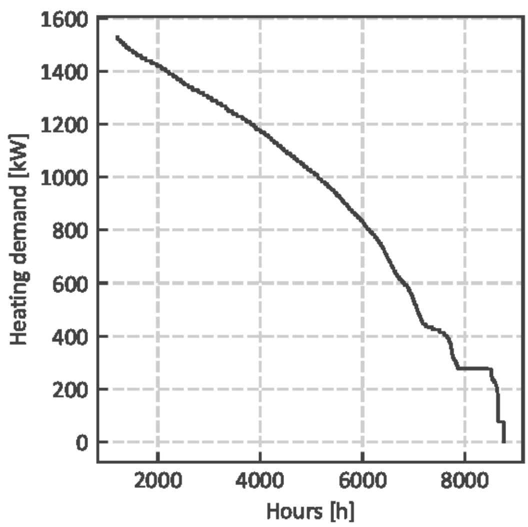

Figure 13: Annual duration line of the heating demand.

7.3 Integration of the CO_2 capture to CHP

Extracted steam at 6.95 bar(a) can be used to supply the desorber of the carbon dioxide removal plant with the required heat. With this, the usage of the produced heat can be increased during the summer. However, this will also lead to a reduction in electricity production of the steam turbine and when the heating demand of the district heating network is high, the heat supply is insufficient to operate both processes in parallel. So the capacity of the carbon capture plant has to be chosen, so that a high amount of full load hours can be obtained while also ensuring economic viability. For the pinch analysis, the operation in the summer is considered, with a full load capacity of the carbon capture plant with a heating demand of 6 MW in the desorber.

Since the required heat for the desorber is at a high temperature level, a proper integration of the process could lead to a high potential for heat recovery and is therefore absolutely essential. For the analysis, the flue gas is only considered after the filter stations. Steam of 6.95 bar(a) and 1.2 bar(a) can be used as hot utility. Some of the streams are not considered in the analysis, because they are assumed as 'enforced matches'. 'Enforced matches' are HEX which cannot be changed due to economic considerations [14] or technical reasons. These are the feedwater heater, where the steam is besides heating the feedwater also necessary for degassing the feedwater. In the partial condenser of the carbon dioxide capture plant, the amine-water solution is condensed. It is assumed that this condenser is also an enforced match. So the cooling water for the condenser is therefore modelled as a replacement stream instead. In addition, the condensate preheater located in the heat distribution network at 95 °C is also assumed to be an enforced match, since the transferred heat in summer operation is rather low. The process data for the streams of the amine treating process used assumes desorber capacity of 6 MW.

7.4 Pinch Analysis

Based on the above consideration the following composite curve is obtained:

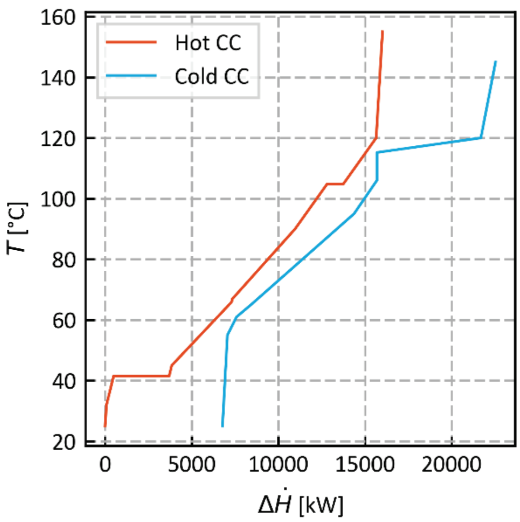

Figure 14: Composite curve of the CHP.

Assuming an overall minimal temperature difference of ΔT_{min} = 9 K, up to 9'191 kW of heat recovery is possible, while 6'567 kW HU and 8'445 kW CU are required according to the composite curve of the process. To size the demand for the Hot Utilities, the grand composite curve is considered Figure 15.

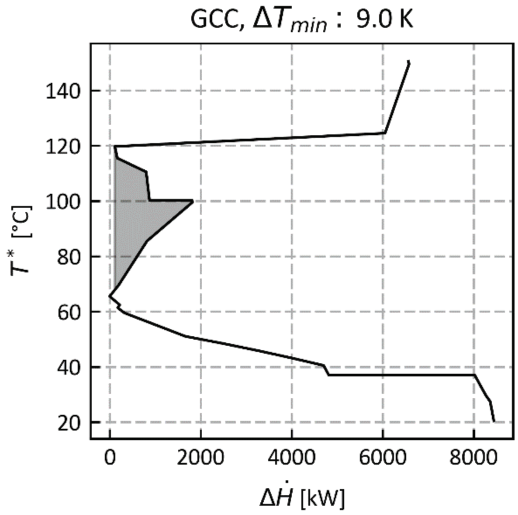

Figure 15: Grand composite curve of the power plant.

As shown in the GCC, there exists a pinch point and a near pinch point in the system with a 'pocket' within the (shifted) temperature range between approx. 65.5 °C and 120 °C. In this range external heating or cooling is not necessary. In the case of the power plant, this means that heating with steam at 1.2 bar(a) (105 °C) from the view of pinch analysis should be avoided.

Above approximately 119 °C, there's a heat deficit, that can only be balanced with hot utility (in this case extracted steam at 6.95 bar(a)). In addition, below 65.5 °C a heat surplus is present. This heat surplus has to be balanced with cold utility. Based on the construction of the CC and GCC, the MER HEN can be created according to the rules of pinch analysis. The aim of the construction of the HEN is to recover as much heat as possible, while also keeping the heat exchanger network as simple as possible.

Table 5: Energy demand and HR potential for MER and relaxed HEN.

HEN	HR (kW)	HU (kW)	CU (kW)	#HEX
MER HEN	9'190	6'566	8'444	18
Relaxed HEN	8'912	6'845	8'296	12

Based on the pinch analysis, the minimal energy requirements are shown in Table 5. In order to reduce the complexity of the HEN, a relaxed version of the HEN is constructed. The design values of the Relaxed HEN are shown in Table 5. Table 6 shows the resulting energy demand in summer operation, (1) without optimization and (2) with optimization. As abovementioned, it is not necessary to use extraction steam to heat the district heating network and the site network operating at 95°C. The pinch analysis shows that a proper integration of the carbon capture plant can reduce the overall heating demand significantly. This means that on the one hand more electricity can be produced and on the other hand the number of full load hours of the carbon capture plant can be increased.

Table 6: Resulting heating and cooling demands before and after applying PA in summer operation.

Optimized process	Heating demand @ 6.95 bar(a) (kW)	Heating demand @ 1.2 bar(a) (kW)	Cooling demand (kW)
Without	7'226	3'132	10'874
With	6'845	-	8'296

Furthermore, static calculations based on the heating demand curves show, that the of full-load hours of the carbon capture plant can be increased from approximately 4'200 h (if it's not integrated) to 5'670 h with an integration as discussed. However, due to the unfavourable temperature level of the condensing amine, the stream can only be used to preheat the district heating network. Exhaust steam is used to provide the required heat, to reach the desired temperature of 95 °C. This needs to be further investigated to ensure optimal control in operation.

8 Technical-economic comparison

In order to make a meaningful comparison of the three CO_2 absorption plant sizes, technical and economic factors are taken into account The technical data are taken from the preceding Section 6. These data serve as direct comparative data in the assessment, as well as input data for economic calculations. An important aspect of the comparison of the different CO_2 capture plant sizes is the total annualized cost (TAC). These costs consist of investment and operating costs. Furthermore, consideration must be given to the potential revenue from the sale of CO_2. For the transport, the CO_2 must be compressed and processed. (Jackson & Brodal, 2018) and (Deng, Roussanaly, & Skaugen, 2019) published the cost estimation for this process regarding the cost estimation for additional infrastructure and the operating costs serve as the basis for the estimates in this work. Items such as international transportation and final disposal are explained in Section 8.1.2.

8.1 Cost calculations for the three CO_2 absorption plants

The absolute and annualized costs for the three CO2 absorption plants, the absolute as well as the annualized investment, operating, and total costs are calculated. The procedure is based on the PhD thesis by Francisco (2021), as well as the paper by (Zanco et al., 2021). Through the comparison with these corresponding data, the calculations can also be validated.

8.1.1 Total annualized investment costs

When considering costs, annualized costs are an important component. The cost analysis by the annuity method shows that for a certain capital recovery factor (ϕ), this is dependent on the CO_2 absorption plant lifetime and a certain interest rate, how high the annual investment costs are. This value corresponds to 0.087. The calculation of the TAC is calculated by this formula:

$$TAC = AIC + AOC$$

The TAC value is made up of the annualized investment costs (AIC) and the annualized operating costs (AOC). The AIC value can be calculated by the total investment costs (TIC). Additional expenses such as installation costs, indirect costs, costs for maintenance buildings, and other additional costs. These are dependent on the total investment costs. The procedure for these calculations is based on the paper mentioned in section 8.1 (Zanco et al., 2021). Due to the identical procedure, the data can be compared and validated.

8.1.2 Extended cost estimation with transport and final storage

In the case of CO_2 absorption, the transport and final storage are the most expensive at the beginning. If infrastructure is created, these costs can be reduced. For this reason, two scenarios are used for the cost calculation, to cover possible future scenarios. These are a short-term scenario (KF) with initially high costs, and a long-term scenario (LF), in which the costs for transport and disposal costs are expected to decrease. Various parameters are adapted or supplemented for economic calculations. For example, the costs for the reboiler steam are 0 CHF. This is because this thermal energy otherwise remains unused. Due to the additional steam extraction, less electrical energy is generated, which is taken into account. The electricity costs for the operation of the CO2 absorption plant and the CO2 compression also amount to 0 CHF. The required electrical energy is subtracted from the annual electrical energy and thus taken into account. By the electricity sales price, the reduced income can be calculated.

Figure 16 presents the annualized investment, annualized operating, transportation, and final storage costs together. In the short-term scenario, the transport and final storage costs are dominant. This is because no infrastructure optimized for CO_2 transport has been created. In the long-term scenario, significantly lower total costs are possible. In this scenario, there is an infrastructure in place for transport and final disposal. According to estimates, transport costs fall from 78 CHF/t CO_2 to 23 - 29 CHF/t CO_2, final disposal costs are expected to decrease from 33 - 61 CHF/t CO2 to 13 - 33 CHF/t CO_2. The total costs decrease from approx. 261 CHF/t CO_2 to 185 CHF/t CO_2. Also, the calculated investment and operating costs offer the potential for improvement. Here, with an optimized/detailed design, costs can still be saved. A possible process integration also offers the potential for savings.

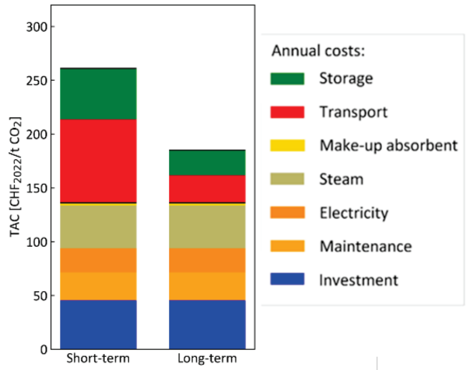

Figure 16: Cost comparison in the short and long term with regard to capture, transport and final disposal.

8.1.3 Sensitivity analysis

With the help of these annualized total costs, the revenues from the sale of CO_2, and the consideration of the revenue shortfall from electricity, possible profit estimations can be carried out. Since, the development of the CO_2 price and the transport and final storage is very uncertain, the sensitivity analysis shows possible scenarios depending on the development of these boundary conditions.

In reality, the CO_2 price and the net revenue from the sale of electricity vary simultaneously. The influence of these changes on the economics of CO_2 absorption is shown in Figure 17. In this graph, the case of 100% flue gas treatment is shown in the case of the short-term (solid) and long-term (dashed) scenario. The highest profit is achieved with a low electricity price, a high CO_2 sales price, and the long-term scenario. With a low CO_2 price, a high electricity price, and the short-term scenario, the lowest revenue is achieved. This is the profitability of the CO_2 absorption plant. If the CHP is also considered, it is negative if the electricity price is too low. The net revenue from the sale of electricity decreases. It is to be expected that there is an optimal case where the captured CO_2 and the electricity can be sold most profitably. From Figure 17, it is possible to estimate how the profit prospects for a given CO_2 price, and net revenue from the sale of electricity. It is also possible to calculate the CO_2 absorption plant differently depending on the CO_2 and electricity prices.

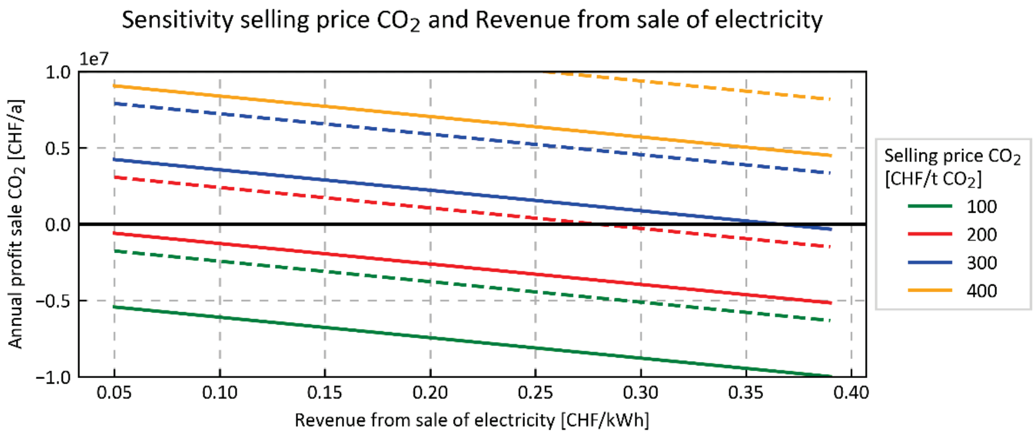

Figure 17: Sensitivity analysis change in CO_2 price and net revenue electricity sales for long-term (solid line) and short-term (dashed) scenarios.

9 Conclusions and next steps

This deliverable report investigates CO_2 capture in a wood-fired power plant using an absorption plant with monoethanolamine as the solvent. The investigation considers techno-economic aspects of the integration and operation of the whole system by understanding how to optimally integrate the CO_2 capture plant into the wood-fired power plant using pinch analysis, with a comprehensive technical-economic evaluation of the process integration options and assessment of different scenarios; in particular, the influence of CO_2 capture on the electricity and heat production of the wood-fired plant was investigated.

The next step from this report is to develop a standard on how to select, design, and optimize an integrated process design for a carbon-capture utilization and storage integrated system in Switzerland, which is a project funded by SFOE, Process-Integrated Carbon Capture: Design and Evaluation (PICC). The scope of this project is to analyze three industrial case studies, where the correlations and differences between the different type of point sources and parameters (operating, model topology, etc.) will be identified. Based on these case studies, a generic guideline for the design and optimization of integrated CCUS systems at a broad range of CO_2 point source types will be developed. To achieve the overall aim a systematic approach (bottom-up) will be taken to address the complexities of designing an integrated CCUS with the case studies. The systematic approach will contain process selection, process integration, flowsheet development, and optimization. These results will inform a top-down analysis by validating and refining a high-level energy system modelling on the impact of CCUS on the Swiss energy system.

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